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OPTIMIZATION OF OIL RECOVERY FROM NEW BRUNSWICK OIL SHALE

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EXECUTIVE SUMMARY

It has been documented by other workers in the field that fast reactors (under one second residence time) tend to produce higher yields of oil from oil shale than do conventional slow reactors (one minute to four hours). These greater oil yields result from reduced coking within the shale particle and less cracking of the oil in the gas phase. An additional benefit of the fast reactors is smaller vessel size and hence capital cost savings.

To confirm these assumptions, a preliminary series of tests was carried out in an ablative fast pyrolysis reactor with New Brunswick oil shale, under the following conditions;

- Temperature: 450 to 650°C
- Residence Time: 400 to 1100 milliseconds
- Particle Size: 375 to 850 microns
- Carrier Gas: Nitrogen and Hydrogen

The results of the program confirmed that the anticipated reactor operating characteristics were, in fact, obtained - a strong dependence of yield on operating temperature and reduced yields for shale with larger particle sizes, especially at the short residence times used in the current reactor configuration.

The tests also successfully demonstrated that, by appropriate selection of operating parameters, the yields could be significantly improved over those obtained in conventional reactors, namely 140% of the Fischer Assay for the ablative reactor versus 90% for the more traditional processes used with this particular shale. The Fischer Assay is a test used to establish the oil yield that could likely be obtained from a conventional retort, not necessarily the maximum oil yield obtainable from the shale.

Reactor conditions in the range of 600 to 650°C and 800 to 1000 milliseconds were necessary to obtain these superior results. The use of atmospheric pressure hydrogen as the carrier gas also produced the same high yields but at lower temperatures, 550°C.

If these improved yields can be maintained in a commercial operating environment, a very elementary economic analysis indicates that a production plant would be economically feasible at today's oil prices ($US 20/bbl). Additional research and economic studies would be necessary to confirm the advantages of introducing this technology for the recovery of oil shale, especially with more productive shales.

Sommaire

Il est couramment admis par les specialistes que les réacteurs rapides (temps de passage inferieur à une seconde) ont tendance à donner des rendements superieur en ce que concerne la production d'huile àpartir de schiste bitumeneux par rapport aux rendements obtenus à partir de réacteurs lents (une minute à 4 heures). Cette performance superieure repose sur une cokefaction reduite dans l'interieur de particule un moins de crackage de l'huile dans la phase (gazeuse). Un avantage additionnel des réacteurs rapides est la possibilité de réduire la taille des réacteurs et consequemment le coût en capital.

Pour confirmer ces affirmations, une série préliminaire de tests a été réalisée à l'aide d'un réacteur rapide pour ablation pyrolytique, à partir schiste bitumeneux dans les conditions suivantes:

- Température: 450 à 550°C
- Temps de passage: 400 à 1100 millisecondes
- Granulometrie: 375 à 850 microns
- Gaz transportation: azote et hydrogène

Les resultats du programme ont confirmé que les performances operationelles prévues du réacteur etaient en fait atteintes, une forte interrelation des rendement avec la température d'operation et une réduction du rendement avec l'utilisation de particules plus grandes, specialement avec les temps de passage courts utilisés dans le cadre des réacteurs courants.

Les tests ont aussi demontré avec succès qu'avec une selection appropriée des paramètres de travail les rendements pourraient être très sensiblement améliores par rapport a ceux obtenus avec des réacteurs conventionnels notamment 140% du Fischer Essai en ce que concerne le réacteur pour ablation par rapport à 90% par le procédés traditionels utilisés pour ce schiste particulier. Le Fischer Essai est un test utilisé pour etablir les rendements un huile qui peuvent être obtenus avec une retorte conventionelle mais pas forcément pour etablir les rendements en huile obtenus à partir du schiste.

Des températures de 600 à 650°C et des vitesses de 800 à 1000 millisecondes etaient indispensables pour obtenir ces resultats. L'utilisation d'une atmosphère d'hydrogène sous pression en tout que gaz porteur permet d'obtenir les même rendements à une température plus basse, à 550°C.

S'il est admis que ces rendements peuvent être également atteind dans un environment operationnel commercial une simple petite analyse économique permettra de savoir si une usine adoptant ce principe serait économiquement viable au cours actuel du pétrole (US 20/Baril). Des études supplementaire seront necessaires pour savoir si l'utilisation d'une telle technologie serait avantageuse dans le cadre de la récupération d'huile de schiste spécialement de schiste plus productif.

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Optimization of Oil Recovery from New Brunswick Oil Shale

1.0 INTRODUCTION

Large deposits of oil shale have been found in the Moncton sub-Basin of New Brunswick. Within this region, the most important deposits are located in the Albert Formation. Oil yields, from the shale in this zone, average as high 100 l/tonne (20 Imp. Gals/ton), as determined by the Fischer Assay. As yet, there are no commercial operations at this site.

Over the past few years, the Research and Productivity Council of New Brunswick has carried out a series of oil production tests, using the shales from the Albert Formation. With a conventional up-draft retort, the group has attained continuous yields as high as 89% of the Fischer Assay, Salib (1984). Subsequent testing by the University of British Columbia, Watkinson (1985), in a spouted bed reactor has produced similar yields. It should be noted that the Fischer Assay is an industry standard test designed to simulate the oil yield that might be expected from a conventional slow retort. The Assay does not measure the ultimate oil yield that is obtainable under optimum pyrolysis conditions and it is possible to obtain oil yields greater than 100% of the Fischer Assay.

The results of these workers are low when compared with shale tests carried out in a fluidized bed reactor by Rubel (1983) at the Kentucky Institute for Mining and Minerals Research. The higher yields attained by this group, 145% of Fischer Assay, have been attributed to the short gas residence time of the fluidized bed system.

Recently, D. B. Brown and J. W. Black developed a novel, yet simple reactor based on an ablation principle which produces very high solid/solid heat transfer rates. As a result,

pyrolysis of solid particles can be accomplished with very short vapour and solid residence times, typically in the range of 50 to 5000 ms. The reactor, which is now patent pending, was developed principally for the surface pyrolysis of organic solids which, when rapidly heated against a hot metal wall, would readily ablate, thus exposing fresh surface for additional decomposition. However the unit will also produce extremely rapid heating rates in solids which do not completely volatilize. It was hypothesized that this reactor could simulate the short vapour residence times and the high oil yields obtained by Rubel (1984) and provide a mechanism for scaling the data to larger systems. Such a system, if successful, could improve the viability of commercial operation from the Albert Formation.

In tests carried out to date with a variety of materials, the behaviour of the reactor has been equivalent to that of other fast pyrolysis reactors, such as the compact fluidized bed developed by University of Waterloo, Scott (1984). For preliminary comparison purposes, it would be reasonable to assume that the information developed in this program would be applicable to other ultra fast reactors.

2.0 OBJECTIVES

The primary objective of the program was to demonstrate that a generic short residence time reactor with ultra rapid heating rates, such as a continuous ablative reactor or a fluidized bed reactor, could produce high yields of oil from New Brunswick shale. Secondarily, the effects of temperature, particle size and residence time were to be investigated to determine the feasibility of such a reactor in this application.

3.0 EQUIPMENT AND PROCEDURES

3.1 Apparatus

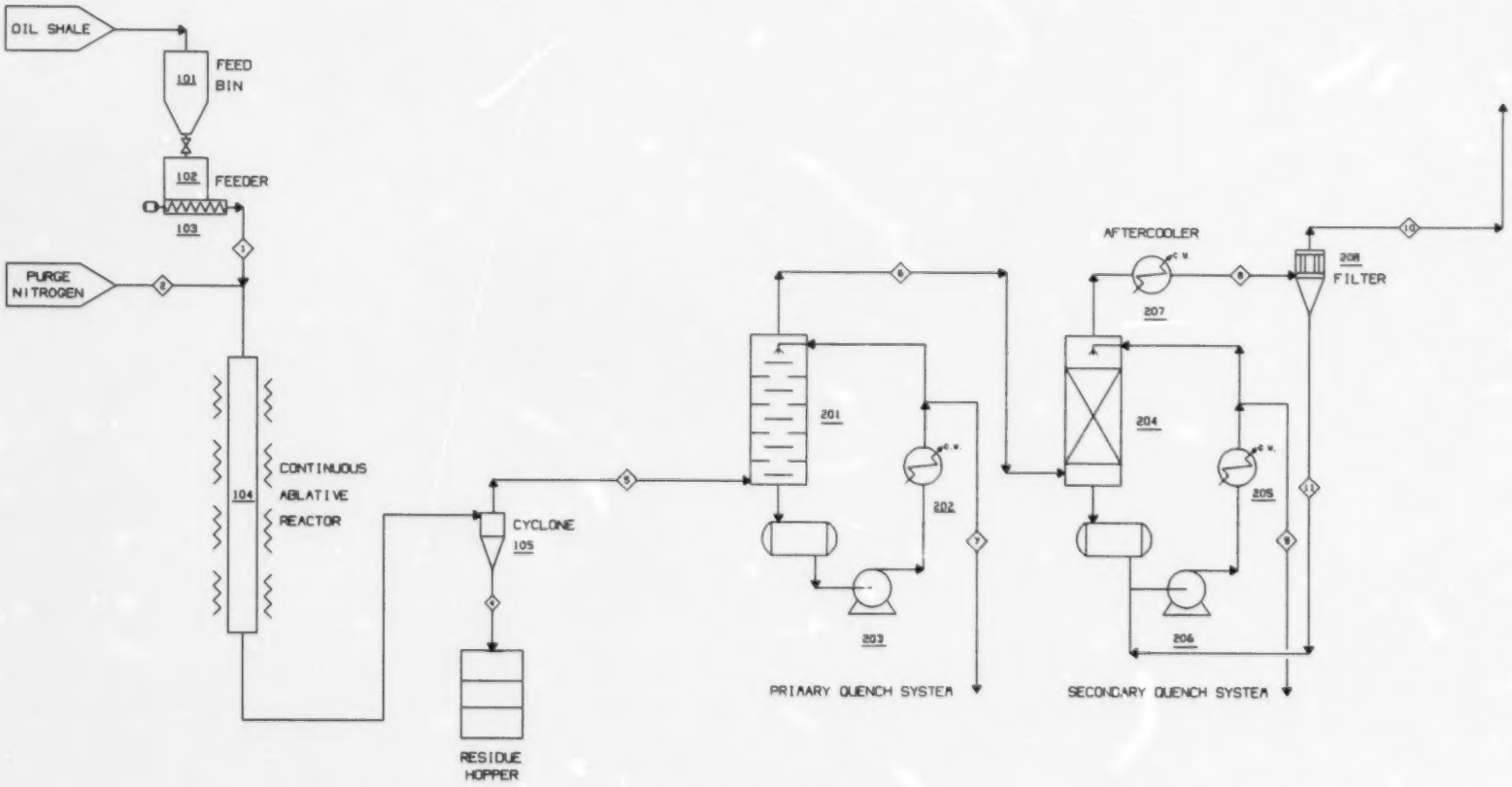
The apparatus used for the tests in this program is shown in Figure 1. The system, which has a capacity of 5 to 20 kg/h, consists of a feed bin mounted on load cells, a continuous ablative reactor (CAR), a cyclone, a condensation train and various measurement equipment.

The ablative reactor used in this project is of novel design and a submission has been filed for Patent protection in a number of jurisdictions. As a result, the detailed design of the reactor cannot yet be made public. It is anticipated that this will occur soon and the details will then be readily available. In the interim, special arrangements can be made to obtain additional information.

In operation, shale was manually loaded into the feed bin, which was then sealed to prevent leakage of gases from the system. The shale was then conveyed from the bin by means of a variable speed discharge screw. Bin weight and hence feed rate were continuously monitored by summing the outputs of three load cells which supported the hopper. After being discharged from the bin, the shale dropped through a flexible chute into a carrier gas to be conveyed into the reactor.

For most of the tests, the carrier gas used was nitrogen. However in one test, hydrogen was employed as the carrier gas. For commercial purposes, the gas produced from the process would be substituted for nitrogen and be recycled to the reactor, by means of a low pressure blower. To minimise the heat load on the reactor, the carrier gas was preheated to the planned operating temperature by a tubular electric heater, prior to mixing with the shale.

FIGURE 1. OIL SHALE PYROLYSIS PILOT PLANT



Both the carrier gas and the shale were injected into the continuous ablative reactor (CAR) where the temperature of the shale was rapidly raised to the desired operating value, by direct intimate contact with a hot metal surface. Six separate heaters, each with its own controller, ensured constant temperature throughout the heating zone where pyrolysis takes place. The products from the reactor were transported into a cyclone where the shale residue was separated from the vapour stream. The residue dropped into a sealed hopper, to be weighed after each test.

The vapours, containing both condensible and non condensible fractions, were piped from the cyclone directly to the first of two stages of direct contact, condensation columns. In these columns, the vapours were cooled against a down-flowing stream of liquor. The liquids were collected in a sump before being pumped back to the column via a water-cooled, annular heat exchanger. The condensed fraction was recirculated with the initial, pre-weighed, liquid charge, until the end of the test, when the total liquor charge was weighed to determine the liquid product yield. By a similar process, additional liquid was condensed in the second condenser, except that the second unit was operated at 40°C rather than the 100°C of the first column.

The gas leaving the second column was piped to a final, indirect, coil heat exchanger before being exhausted through a mist filter. The condensate from this column was also weighed after each test.

A small side-stream flow was removed from the main product line just after the cyclone and passed through a hot gas filter, an ice-cooled condenser, a mist filter and a dry gas meter, before being vented to the atmosphere. The condenser and mist filter were washed with methylene chloride after each test, for subsequent recovery and weighing. The clean gas was injected into an on-line Carle gas chromatograph containing an FID

and a TCD to determine both the permanent gases and the hydrocarbons up to C_3. Vapours with a carbon atom content more than five were separated as a single product, C_6^+.

Pressure and temperature were measured throughout the apparatus. These measurements, together with load cell and flow rate data, were transferred to a Hewlett Packard data acquisition system, controlled by a computer for instantaneous readout and storage.

3.2 Raw Material Properties

A 1 tonne sample of oil shale was excavated from the Albert Mines zone, 25 km south east of Moncton, by Three-D Geoconsultants Limited. This location is the same zone as that from which samples were taken by Salib (1984) and by Watkinson (1985). The material which was removed by backhoe, was transported to the laboratories of the Research and Productivity Council (RPC) in Fredericton. There, it was crushed by hammermill, dried and screened. The composite sample analysis is provided in Table I, together with the analyses of Salib (1984) and Watkinson (1985). The screen analysis is included in Table II.

Samples of each screen fraction were analyzed by RPC to determine organic and inorganic carbon as well as to measure the oil yield by Fischer Assay. These data are shown in Table II and Figure 2. In general, it can be seen that as the particle size diminishes both the organic carbon content and the oil yield decrease. This observation is similar to that noted by Watkinson (1985). From the more detailed discussion of the oil yield in appendix 1, there is a reasonable probability that the absolute oil yield values for particle sizes, +35 mesh and +60 mesh, are a little low and high respectively.

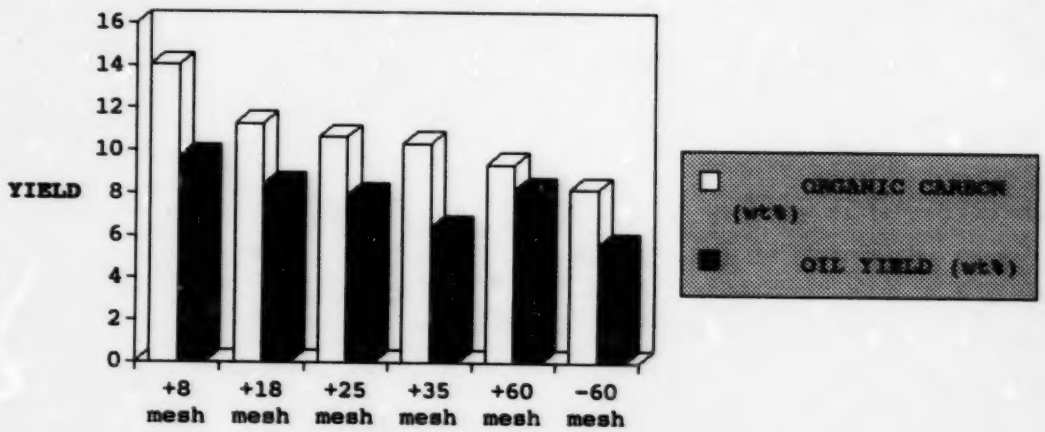
TABLE I- PROPERTIES OF COMPOSITE SHALE SAMPLES

TEST SAMPLE	BBC ENGINEERING	U. of BRITISH COLUMBIA	RES. & PROD. COUNCIL N.B.
FISHER ASSAY			
OIL YIELD (%)	7.23	7.13	7.76
WATER YIELD (%)	2.08	2.00	1.80
GAS LOSS (%)	2.36	2.89	1.14
ORGANIC CARBON (%)	9.86	11.55	11.40
INORGANIC CARBON (%)	3.05	2.39	3.36
ULTIMATE ANALYSIS (Dry basis)			
% C	12.91	15.91	14.76
% H	---	2.05	1.86
% N	---	0.51	0.10
% S	---	0.92	0.77
% Cl	---	0.01	---
% Ash	75.84	73.78	77.55
% O (By difference)	---	6.82	4.96
ASH ANALYSIS			
SiO2 (wt %)	---	42.30	44.84
Al2O3 (wt %)	---	10.40	12.11
Fe2O3 (wt %)	---	4.43	6.10
CaO (wt %)	---	8.45	7.88
MgO (wt %)	---	3.38	3.17
Na2O (wt %)	---	1.02	1.64
K2O (wt %)	---	1.60	1.80
SO3 (wt %)	---	1.93	---
Balance (wt %)	---	0.27	---
TOTAL	---	73.78	77.54

TABLE II- EFFECT OF SIZE ON SHALE PROPERTIES

SIEVE FRACTION	+8 mesh	+18 mesh	+25 mesh	+35 mesh	+60 mesh	-60 mesh
FRACTION PROPERTIES						
RETAINED (wt%)	2.7	27.2	11.2	9.0	16.5	33.4
MOISTURE CONTENT (wt%)	1.33	1.62	1.36	1.41	1.65	---
ORGANIC CARBON (wt%)	14.05	11.28	10.65	10.35	9.38	8.20
INORGANIC CARBON (wt%)	5.44	4.32	2.63	2.94	3.50	1.76
ASH CONTENT (wt%)	77.69	76.67	74.41	75.52	75.32	---
FISHER ASSAY						
OIL YIELD (wt%)	9.70	8.40	7.90	6.40	8.20	5.60
WATER YIELD (wt%)	2.00	2.20	2.10	1.80	2.20	2.00
GAS LOSS (wt%)	2.30	1.80	2.20	3.70	1.80	2.80
TOTAL	14.00	12.40	12.20	11.90	12.20	10.40

FIGURE 2 - EFFECT OF PARTICLE SIZE ON YIELD



4.0 RESULTS

The detailed results are presented in Appendix 2, while selected results are provided in Tables III, IV and V.

Two problems were encountered during the program which resulted in some difficulty in directly closing the mass balance for some tests. The first of these involved the free flowing nature of the shale used in the tests. As a result of this high flowability, material would occasionally spill from the bin into the conveying line after the feed screw was stopped, at the beginning or at the end of a test. Thus the weight of the feed and the residue were sometimes in error. Quick fixes were unsuccessful and the problem was only resolved when a new short pitch screw was installed. Mass balances, after the modification was carried out (runs 18 and later), were within 5% of theoretical. Unfortunately, there was insufficient shale to duplicate the initial tests.

Secondly, because of the small amount of oil produced in each experiment, in comparison with the initial oil charge, errors were easily introduced in measuring the liquids drained from the condensers at the end of each test. This was particularly evident with high temperature runs which produced high viscosity liquids which did not drain well.

To ensure consistency of the mass balance, the final closure, as shown in Tables III to V, was calculated by the following procedure:

Shale weight:- The load cell readings were statistically averaged over the flow period to obtain the mean flow rate which was then multiplied by the run time to obtain the total input weight.

Residue weight:- A composite sample of the residue was obtained from a series of thieved samples from the residue hopper. The composite was thoroughly mixed and ashed in duplicate. The residue weight was calculated from the ratio of ash in the residue to ash in the raw shale. The latter is shown

TABLE III- RESULTS (375 Micron Particles)

RUN NUMBER	26	27	28	29	30	34	1	33
REACTOR PARAMETERS								
PARTICLE SIZE (microns)	600	600	600	600	600	600	1690	600
TEMPERATURE oC	550	550	500	600	650	550	500	550
CARRIER GAS	N2	N2	N2	N2	N2	N2	N2	H2
REACTOR PRESSURE (kPa)	112	112	125	125	134	125	129	108
RESIDENCE TIME (ms)	713	705	895	802	828	1094	435	318
SOLIDS FLOW RATE (kg/h)	7.3	7.3	6.5	6.5	6.5	7.2	15.9	6.1
MASS BALANCE								
(Normalized at 100kg)								
Shale (kg)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Shale Residue (kg)	92.15	92.56	96.83	89.00	88.45	89.55	98.41	90.45
Liquids (kg)	6.42	4.82	2.93	8.18	8.78	6.13	0.47	8.89
Gas (kg)	2.10	1.99	0.80	3.85	7.41	4.58	0.76	4.41
Water Vapour (kg)	0.35	0.53	0.89	0.39	1.58	0.69	0.27	3.10
Unaccounted (kg)	-1.02	0.10	-1.45	-1.41	-6.22	-0.95	0.10	-6.85
GAS COMPOSITION (kg)								
H2	---	---	0.000	0.003	0.029	0.079	0.004	n.d.
CO2	---	---	0.227	0.723	1.394	0.522	0.268	0.021
CO	---	---	n.d.	n.d.	0.102	0.052	n.d.	0.097
CH4	0.652	0.442	0.111	0.650	1.144	0.687	0.062	0.954
C2H4+C2H6	0.898	0.382	0.197	1.226	2.111	1.456	0.260	1.833
C3H6	0.163	0.231	0.120	0.507	1.246	0.928	0.086	0.923
C3H8	0.225	0.228	0.016	0.109	0.148	0.084	0.004	0.136
C4H6	0.050	0.115	0.021	0.094	0.276	0.201	0.025	0.128
C4H8	0.067	0.352	0.058	0.244	0.473	0.332	0.035	0.248
C4H10	0.008	0.124	0.007	0.044	0.279	0.030	0.011	0.030
C5H10	0.023	0.043	0.032	0.151	0.154	0.114	0.000	0.009
C5H12	0.010	0.070	0.012	0.093	0.057	0.077	0.002	0.032
TOTAL	2.097	1.988	0.803	3.844	7.413	4.562	0.756	4.410

TABLE IV- RESULTS (850 Micron Particles)

RUN NUMBER	18	19	20	21	22	23	24	25
REACTOR PARAMETERS								
PARTICLE SIZE (microns)	850	850	850	850	850	850	850	850
TEMPERATURE oC	500	550	600	650	450	550	550	550
CARRIER GAS	N2	N2	N2	N2	N2	N2	N2	N2
REACTOR PRESSURE (kPa)	122	119	122	136	119	122	129	112
RESIDENCE TIME (ms)	851	748	759	802	877	536	407	1000
SOLIDS FLOW RATE (kg/h)	8.61	6.91	5.18	5.61	5.22	7.88	7.88	7.6
MASS BALANCE								
(Normalized at 100kg)								
Shale (kg)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Shale Residue (kg)	94.88	91.29	87.43	86.66	97.10	93.68	93.62	90.93
Liquids (kg)	1.17	3.86	6.83	3.30	1.76	4.48	3.03	2.20
Gas (kg)	0.76	2.84	2.88	---	0.28	1.06	2.29	1.26
Water Vapour (kg)	0.52	0.31	0.66	1.82	1.45	1.73	1.07	0.50
Unaccounted (kg)	2.67	1.70	2.21	8.23	-0.59	-0.94	-0.01	5.11
GAS COMPOSITION (kg)								
H2	n.d.	0.012	0.013	---	n.d.	n.d.	0.003	0.005
CO2	0.099	0.688	1.030	---	0.159	0.205	0.368	---
CO	n.d.	n.d.	n.d.	---	n.d.	0.001	n.d.	n.d.
CH4	0.119	0.410	0.338	---	0.024	0.231	0.325	0.437
C2H4+C2H6	0.102	0.879	0.692	---	0.039	0.333	0.909	0.376
C3H6	0.051	0.405	0.394	---	0.020	0.118	0.318	0.164
C3H8	0.334	0.069	0.064	---	0.005	0.028	0.031	0.175
C4H6	0.011	0.117	0.111	---	0.006	0.021	0.069	0.034
C4H8	0.030	0.201	0.152	---	0.020	0.062	0.189	0.047
C4H10	0.003	0.016	0.014	---	0.000	0.011	0.010	0.004
C5H10	0.008	0.034	0.043	---	0.004	0.031	0.044	0.010
C5H12	0.004	0.014	0.025	---	0.000	0.016	0.022	0.006
TOTAL	0.761	2.843	2.875	0.000	0.277	1.058	2.289	1.257

TABLE V- RESULTS (600 and 1690 Micron Particles)

RUN NUMBER	7	12	13	14	15	16	17
REACTOR PARAMETERS							
PARTICLE SIZE (microns)	375	375	375	375	375	375	375
TEMPERATURE oC	550	600	500	550	450	650	500
CARRIER GAS	N2	N2	N2	N2	N2	N2	N2
REACTOR PRESSURE (kPa)	114	126	139	136	143	129	134
RESIDENCE TIME (ms)	673	819	1077	893	1160	811	445
SOLIDS FLOW RATE (kg/h)	7.11	6.29	6.08	5.90	5.83	6.11	4.81
MASS BALANCE							
(Normalized at 100kg)							
Shale (kg)	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Shale Residue (kg)	90.47	88.50	96.42	92.38	97.65	87.65	97.06
Liquids (kg)	4.58	5.69	3.44	6.91	3.42	8.74	2.51
Gas (kg)	2.04	2.81	0.76	1.91	0.36	5.39	0.99
Water Vapour (kg)	0.86	0.40	0.64	0.28	0.52	1.88	0.90
Unaccounted (kg)	2.05	2.61	-1.25	-1.47	-1.95	-3.65	-1.46
GAS COMPOSITION (kg)							
H2	0.019	0.004	0.001	0.009	n.d.	0.004	n.d.
CO2	0.495	0.729	0.249	0.554	0.191	0.806	0.541
CO	0.025	n.d.	n.d.	0.016	n.d.	n.d.	n.d.
CH4	0.303	0.397	0.114	0.240	0.030	0.820	0.047
C2H4+C2H6	0.562	0.851	0.151	0.473	0.066	2.043	0.078
C3H6	0.277	0.381	0.068	0.251	0.027	0.851	0.198
C3H8	0.163	0.063	0.017	0.065	0.022	0.095	0.037
C4H6	0.065	0.088	0.081	0.046	0.006	0.189	0.006
C4H8	0.090	0.176	0.049	0.138	0.011	0.345	0.054
C4H10	0.007	0.022	0.008	0.028	0.002	0.030	0.014
C5H10	0.014	0.065	0.013	0.055	0.005	0.109	0.011
C5H12	0.015	0.031	0.011	0.037	0.003	0.044	0.007
TOTAL	2.036	2.806	0.762	1.913	0.363	5.336	0.994

in Table II.

Mass of gas:- The gas produced was calculated from the total nitrogen flow and the average gas analysis measured by the on-line gas chromatograph.

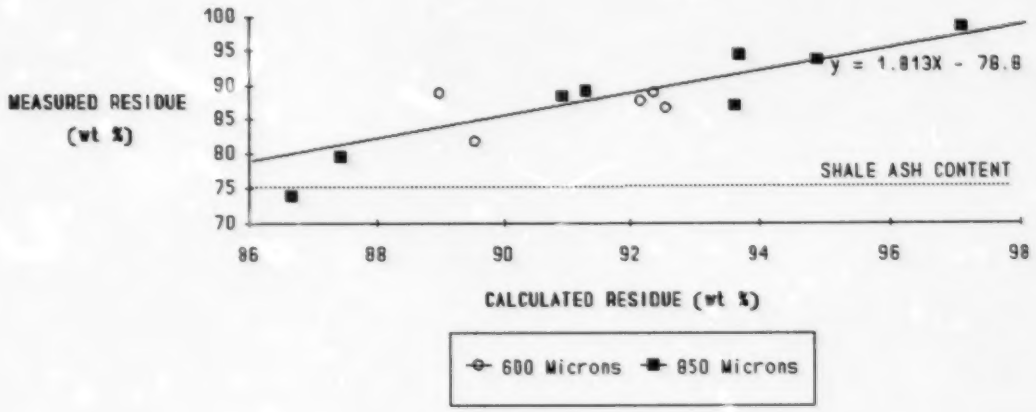
Liquid yield:- The methylene chloride washings from the gas sample line were evaporated in a Kuderna-Danish flask to remove the solvent. The resulting shale oil was weighed. The C_6+ fraction, measured by the gas chromatograph, was added to the condensed liquid to calculate the total sample-line oil yield. In normal operation with recycled gas, these vapours would condense as part of the oil stream. The total liquid yield was calculated by the same method used by Watkinson (1985), namely

$$\text{Oil Yield} = \text{Sample Line Oil Yield} \times \frac{\text{Total Gas Flow (process)}}{\text{Total Gas Flow (sample)}}$$

The information calculated by this procedure was used to evaluate the mass balance, normalized to 100kg, as shown in the tables of results. Two test runs were eliminated from the data shown in the appendix, because the calculated mass balance exceeded the $\pm 10\%$ criterion for acceptance. One test, run 33, had hydrogen as the carrier gas.

The residue yields, as calculated above, were compared with the measured values for runs 18 through 34, as is illustrated in Figure 3. Except for runs in which the volatile release was small, the calculated residue yield was consistently higher than the measured data. Although no conclusive proof was obtained directly, it did appear that the residue, subsequent to the test, either adsorbed or reacted with a species such as oxygen, carbon dioxide or water. All samples were dried at 108°C, prior to ashing. It was also observed that in the mass balance results of Salib (1984), the product yield was always in excess of the initial reactant weight, +4.6 to +14.3%, for all of the direct pyrolysis tests. Yet in the partial oxidation tests, the output mass was consistently lower than the raw material weight.

FIGURE 3 - CALCULATED VERSUS MEASURED RESIDUE YIELDS



Since no concrete explanation for the phenomenon was identified, the residue values were not further adjusted. The calculated residue yields in Tables III to V are probably high for the runs which produced good conversion. From the perspective of the program, the actual values of residue yield are relatively unimportant.

5.0 DISCUSSION

Fast pyrolysis reactors, such as the entrained bed, impinging solids or ablative reactors, provide short residence times (less than 1 second) for both the solid and vapour phases. As a result, these reactors perform quite differently from the more conventional reactors like the fixed bed retort, the rotary kiln or even the fluidized bed, all of which have solids residence times ranging from 20 minutes to 4 hours.

With long residence times, oil can be released slowly from the shale at temperatures of 450 to 500°C. When the transformation time is short, the operating temperature must be raised to achieve the same yield, for two reasons;

- A finite time is required to heat the interior of the particle and this time becomes longer with increasing particle size. To accomplish the heating of the entire particle within one second or less necessitates the use of higher temperature differentials, to overcome the low thermal conductivity of the solid.
- Conversion of the kerogen - bitumen phase to oil is a decomposition reaction with first order kinetics and an Arrhenius activation energy of 120 to 160 kJ/mole. To produce the fast decomposition rates necessary to compensate for the short residence times, it is essential to increase the reaction rate by raising the temperature of the particle.

In addition to the above, it appears that the more rapidly that the oil can be released from within the shale particle, the less likely it is that the oil and the kerogen will undergo transformation to coke inside the shale. Thus the maximum oil yield will probably be higher in rapidly heated systems. Conversely, the oil vapours will be exposed to higher temperatures and may be subjected to more cracking, with additional gas formation. It would be therefore expected that, for fast reactors, oil production will increase with temperature to a maximum and then decrease, while gas yield will continue to grow.

Residence time and particle size will also be significant factors in determining reactor performance and oil yield. Thus all three parameters were investigated in this program. Also, the effect of a reactive carrier gas was examined in one test.

5.1 Effect of Temperature

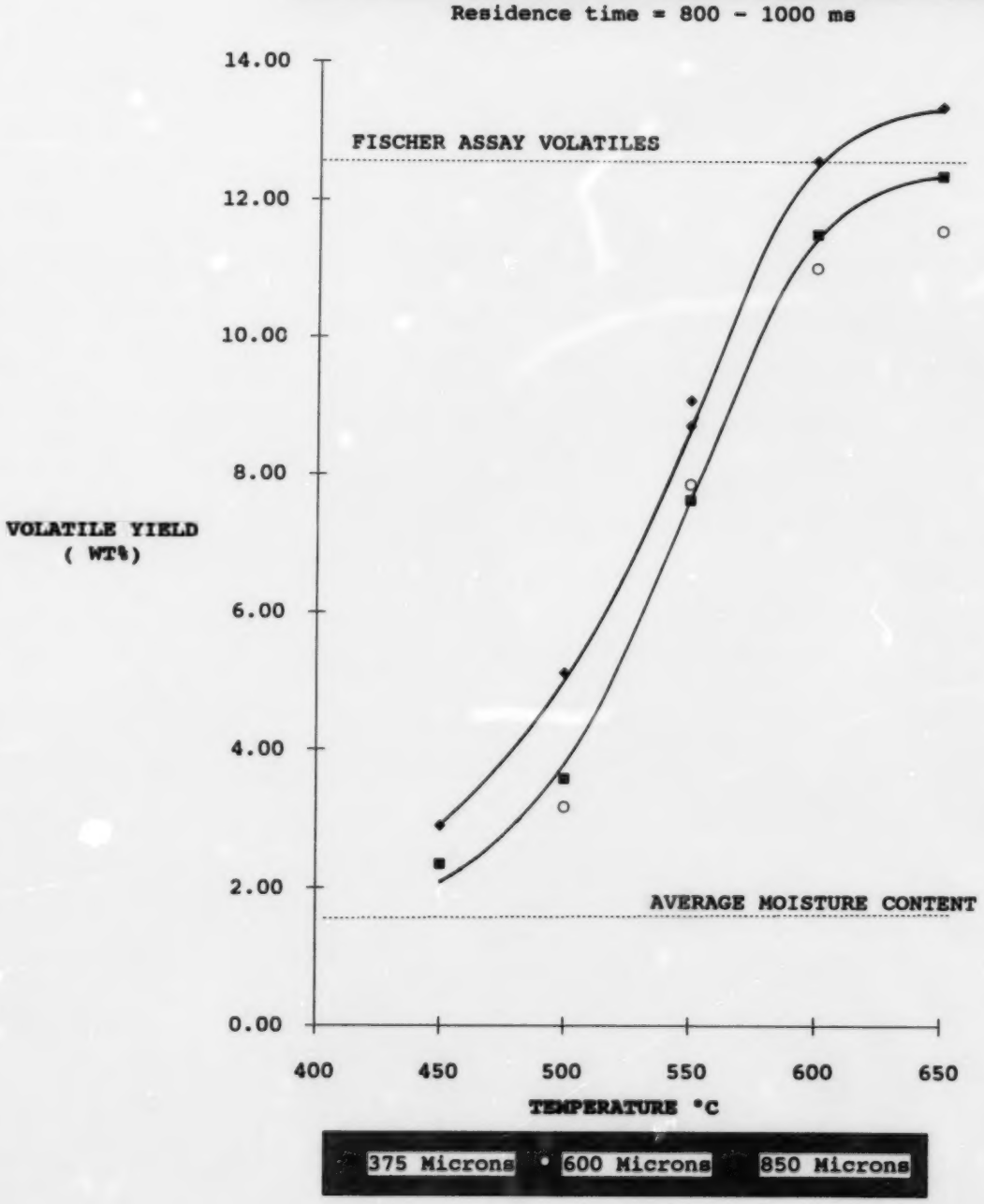
As the shale particle heats up, it will initially lose moisture. With continuing rising temperature, the kerogen in the shale will begin to decompose with the production of gas and oil. Further heating and/or longer soaking will produce more oil, until all of the volatiles are released from the shale. In addition to decomposition of the kerogen, inorganic carbonates present in the shale will also break down to produce carbon dioxide, usually however, at higher temperatures.

In keeping with the nature of reactor behaviour, as described above, and with solids residence times limited to about one second, it should be obvious that operating temperature will play a major role in the overall production of volatiles from the shale. This is, in fact, confirmed in Figure 4 where the total volatiles, calculated from the residue weight loss measurements, are shown to increase with temperature.

In spite of the fact that the weight loss values may be under-estimated, section 4.0, total volatiles (gas, water and oil) are higher with this type of reactor than might be predicted by the Fischer Assay. However, to obtain these results it is necessary to operate the reactor at temperatures higher than the 500 to 550 °C used in conventional retorts.

An examination of Figure 4 indicates that the volatile yield rises rapidly at 400°C from a value of about 1.5%, the initial moisture content, to almost 600°C. At this temperature, the yield begins to flatten out as the reaction approaches completion.

FIGURE 4 - EFFECT OF TEMPERATURE ON VOLATILE YIELD



It would appear from the data, that the 600 and the 375 micron particles produce essentially the same results (the 600 micron particles have slightly less volatiles than the 375 and 850 micron particles). The 850 micron particles consistently appear to produce slightly more volatiles than the other particle sizes, by about 8%. Whether this is due to a particle size effect or to small differences in chemical composition would require further experimentation to determine.

A more detailed picture of the influence of temperature can be had from an examination of Figures 5 and 6. Oil production follows a similar pattern to the volatile yield described above, namely, a rapid rise from 500°C to 600°C followed by a levelling off above 600°C. The values measured are substantially in excess of the Fischer Assay yields, particularly for the 600 micron material which was 40% above the Assay value. To attain these high levels of production, at the residence times tested (approximately 800 ms), it is necessary to operate at temperatures exceeding 600°C.

Oil yields from the 375 and the 600 micron samples are essentially the same, in spite of the fact that the larger particles apparently contain less oil. The largest particles, 850 microns, show consistently lower yields, probably because the finite time required to heat the particle and overcome the kinetic limitations for this particle size, is longer than the time that the current configuration provides.

Gas yields also increase quite rapidly from 500°C. However, they do not tail away at the temperatures tested in the present program. Highest gas production is given by the 600 micron material.

Individual gas yields, Figures 7 to 9, pattern the overall gas production. Methane and propylene are produced in almost the same amounts, irrespective of temperature and particle size. Carbon monoxide and hydrogen are negligible except at high tempera

FIGURE 5 - EFFECT OF TEMPERATURE ON OIL YIELD

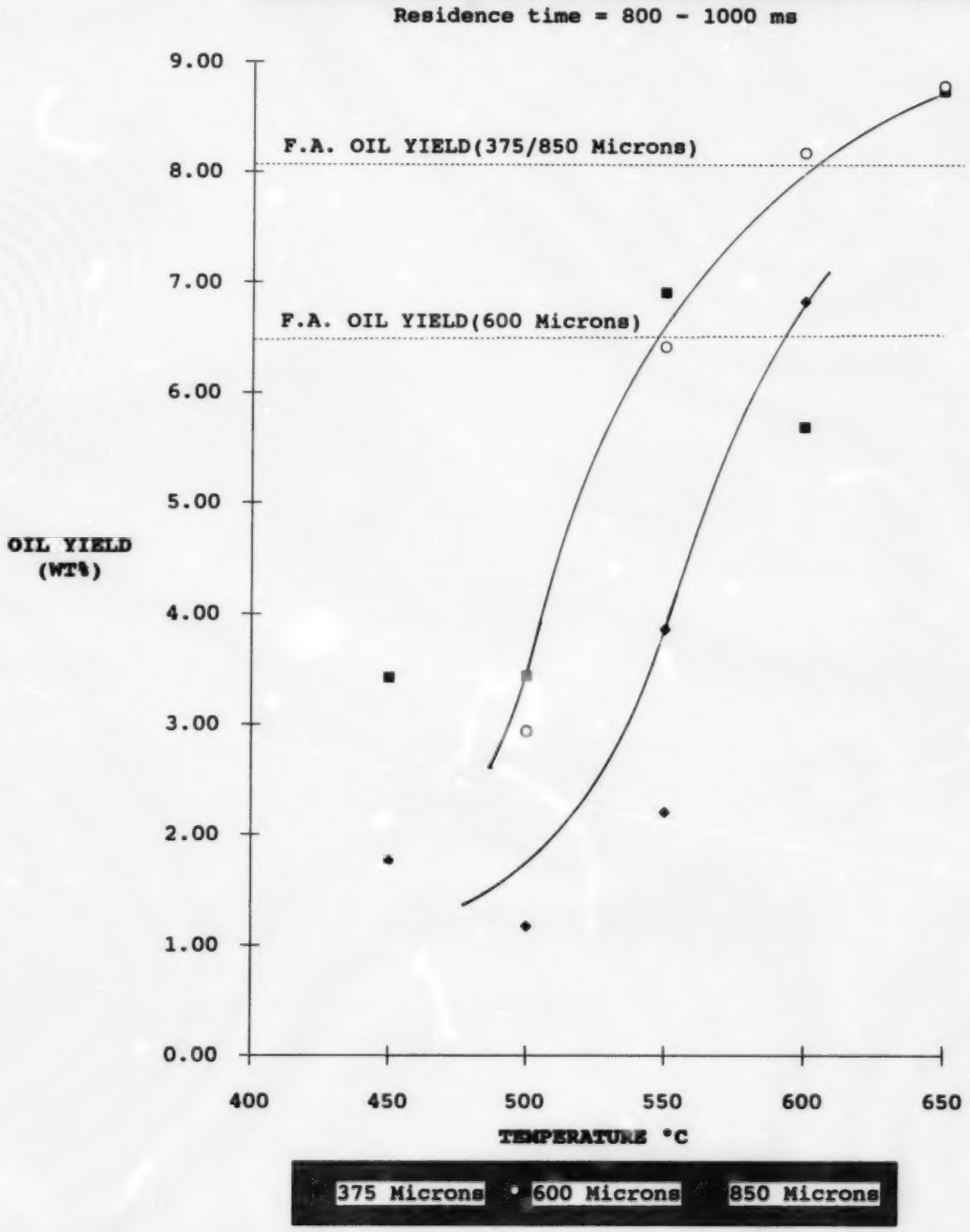


FIGURE 6 - EFFECT OF TEMPERATURE ON GAS YIELD

Residence time = 800 - 1000 ms

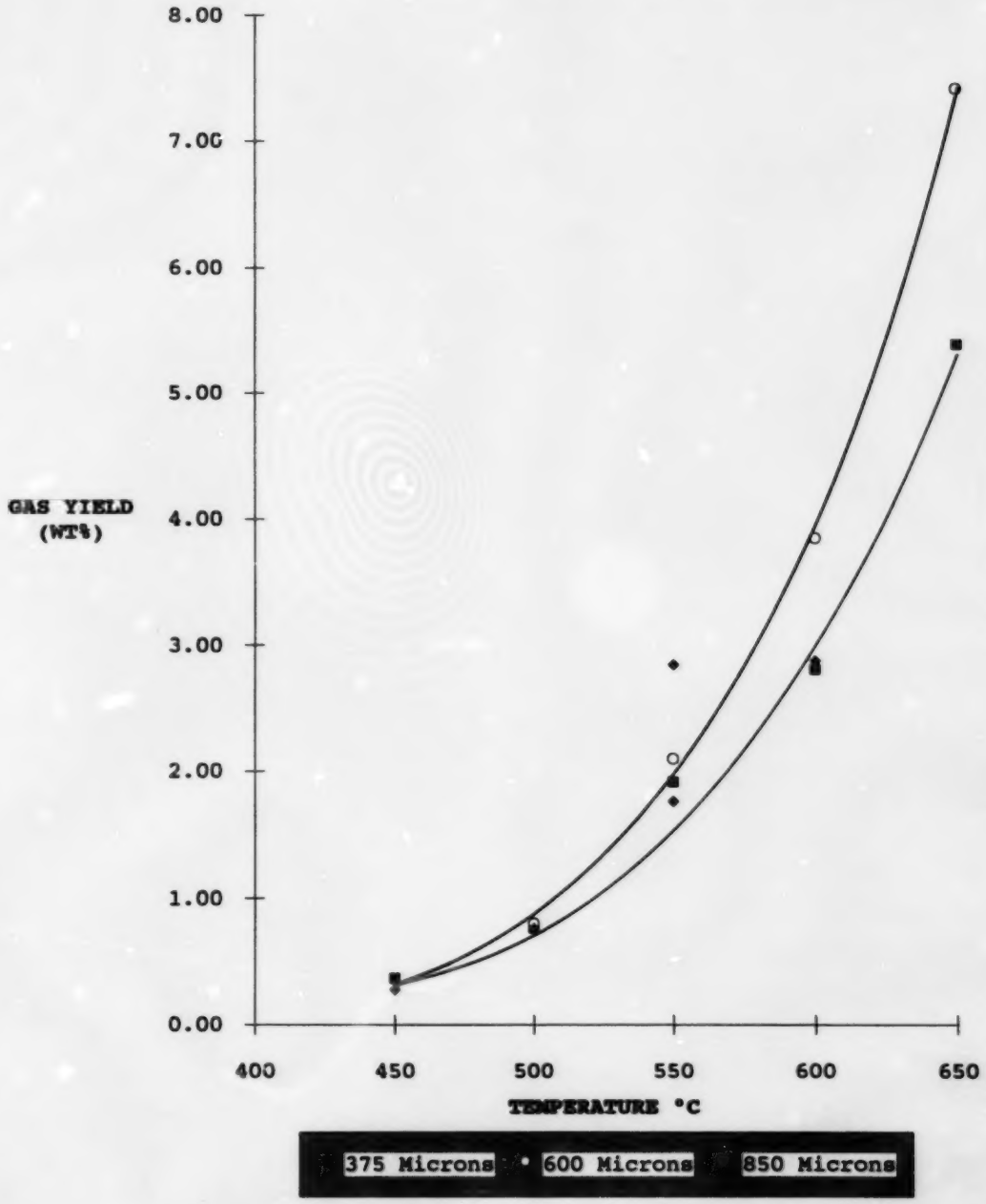


FIGURE 7 - EFFECT OF TEMPERATURE ON CO2 YIELD

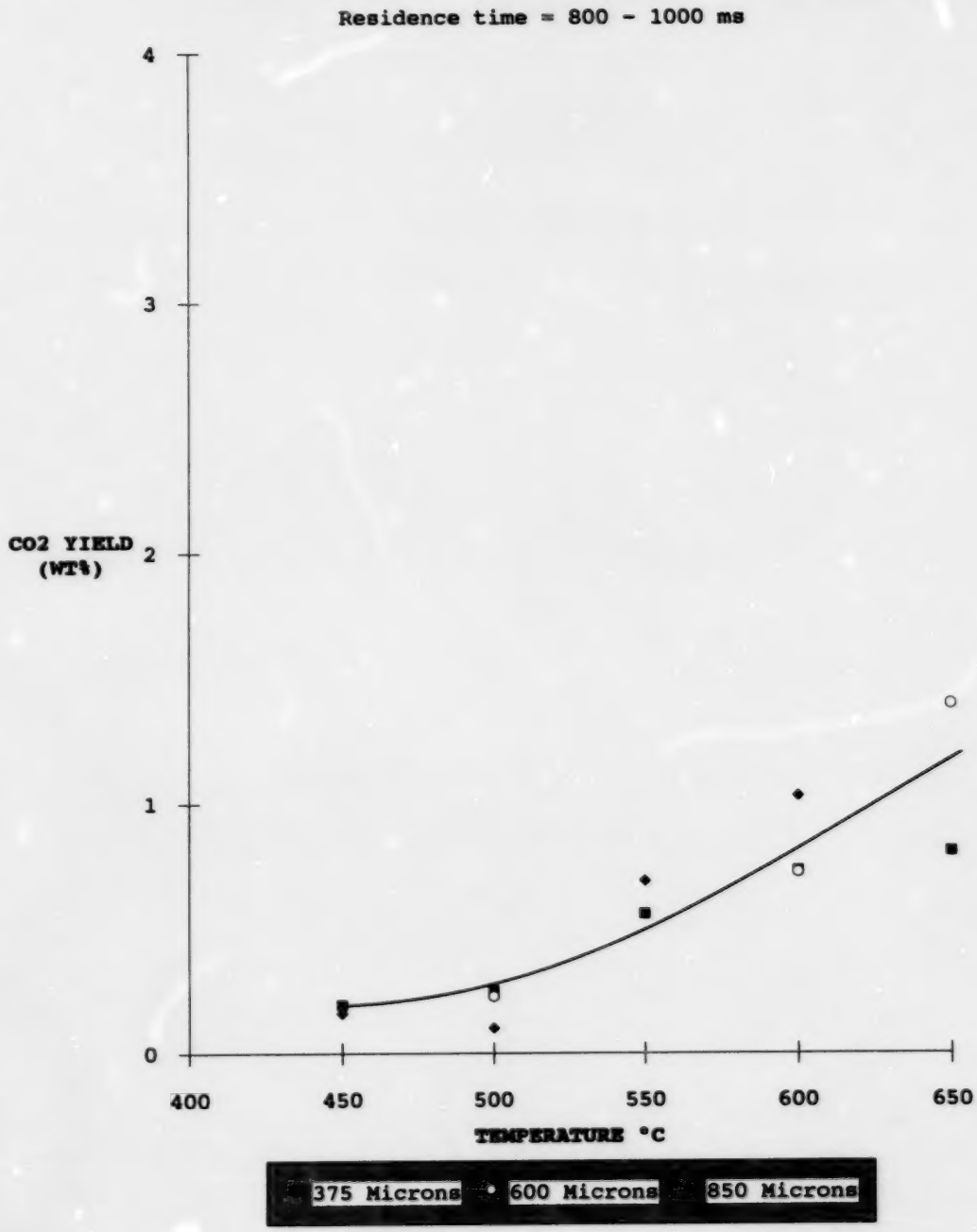


FIGURE 8 - EFFECT OF TEMPERATURE ON CH4 YIELD

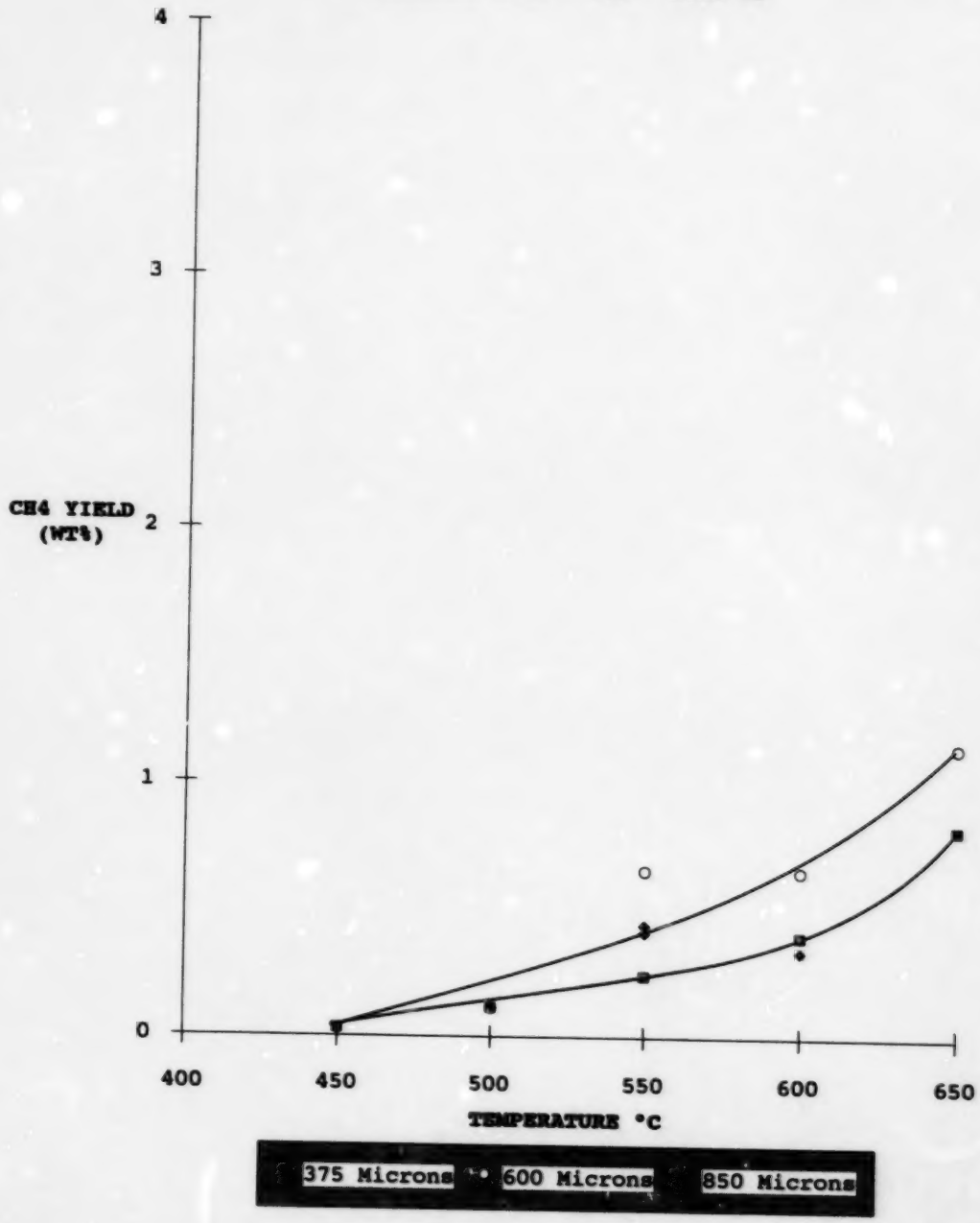
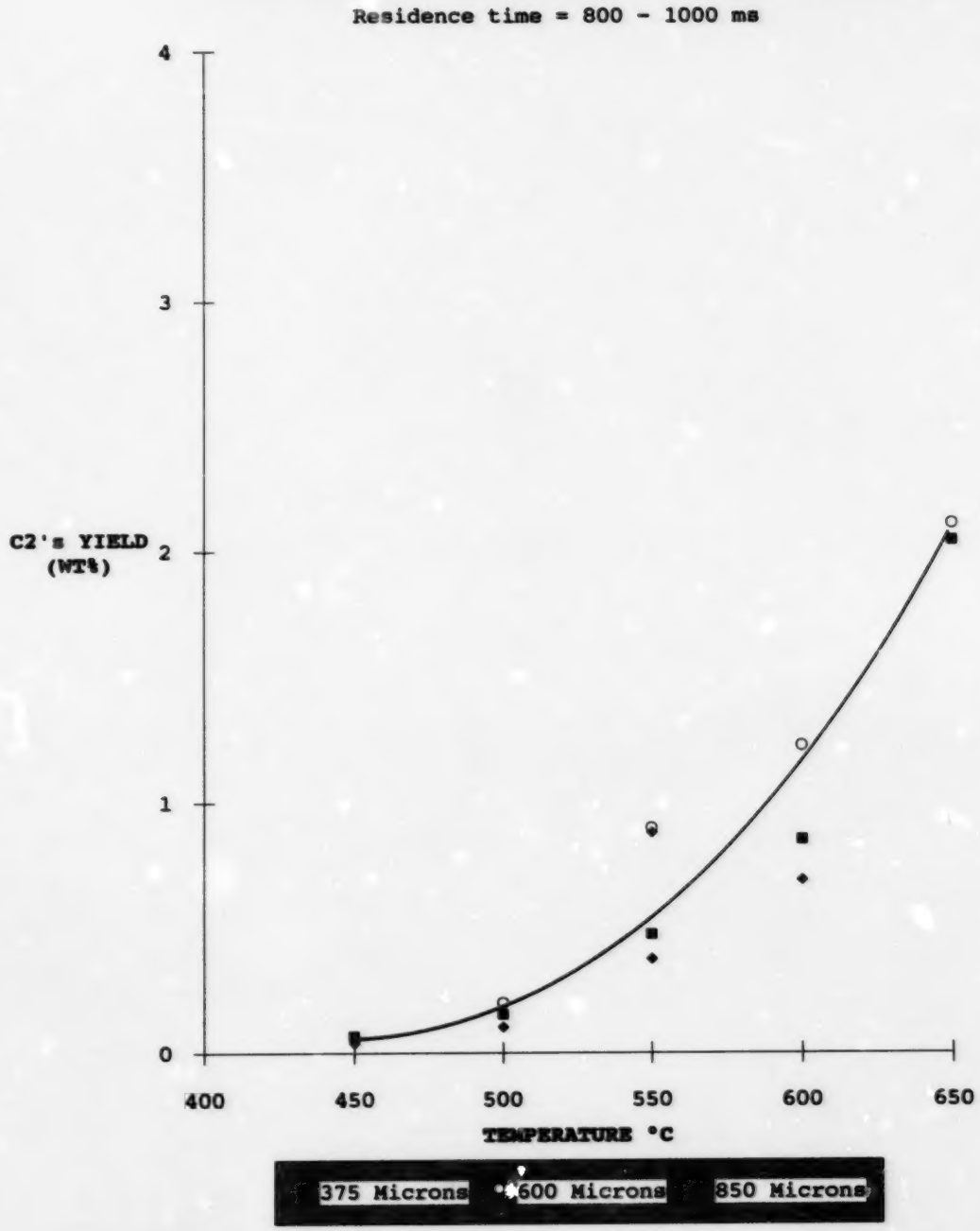


FIGURE 9 - EFFECT OF TEMPERATURE ON C2's YIELD



ture, when the yield does increase but both gases are still insignificant. The most important product is the ethylene/ethane component, of which the ethylene comprises 75 to 80% of the fraction.

Carbon dioxide is only partially released from the shale under the conditions used in this test program. This observation is similar to that of spouted bed tests of Wilkinson. The work of Salib, on the other hand, showed much higher yields of total gas, hydrogen, carbon monoxide, and carbon dioxide. In fact, the gas yields of Salib, were almost twice the liquid yields in some instances, probably as a result of the longer solids residence times, about 4 hours, and slower heating rates.

It can be concluded that, because of the short residence times available to the shale particle, reaction kinetics and particle heating rates play a dominant role in the oil output. It is also possible that even modestly higher yields may be obtained by optimising particle size, residence and temperature.

5.2 Effect of Residence Time

As discussed earlier, a finite time is required to heat the kerogen to the necessary decomposition temperature and subsequently to convert the kerogen into oil. At low temperatures, 450 to 500°C, the time required to complete the conversion, typically a few minutes, is beyond the limits of short solids residence time reactors, such as the entrained bed or ablative reactors. As the operating temperature increases and/or the particle size decreases, the time required to decompose all of the kerogen is reduced to one second or less, which is now within the purview of the ablative reactor. If the

temperature becomes too high or the residence time too long, the oil will tend to crack and the resultant oil yield will begin to fall off.

A number of tests, with residence times ranging from 400 to 1100 ms, were carried out at 550°C. Unfortunately the data on oil and gas yield contain sufficient scatter that the differences resulting from longer residence time are somewhat masked. However, the measurements based on total volatile yield, as determined from the residue analysis, do show an increase in volatile production with time, Figure 10. These data can be correlated by a linear equation;

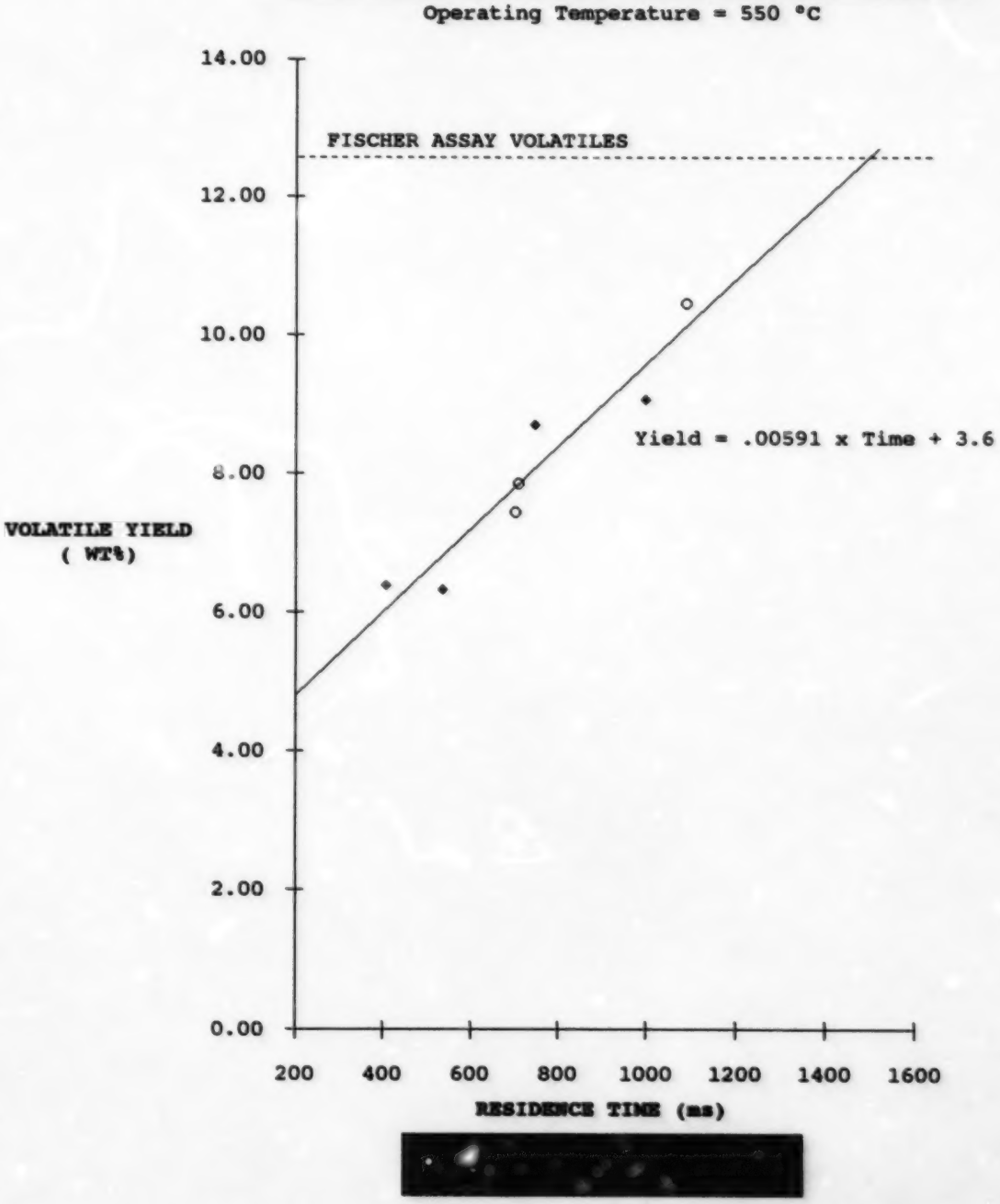
$$\text{Yield} = 0.00591 \times \text{Residence Time} + 3.64$$

The correlation coefficient, r^2, is equal to 0.91, indicating a good fit over the range of residence time tested. If the correlation is extrapolated to the time necessary to attain the Fischer Assay yield at 550°C, it would appear that a residence time of 1500 to 1600 ms is required, about 50% more than is available from the current pilot plant configuration. With a modest increase in temperature, it would be expected that essentially complete conversion could be attained within the residence time of the existing reactor. Indeed, this is confirmed in Figure 4, where actual Fischer Assay yields were obtained at 600°C.

5.3 Effect of Particle Size

The analysis of particle size effects is complicated by the fact that the properties of each of the size ranges tested were different, particularly for Fischer Assay and organic carbon content. As described in section 5.1, it appears that particles 850 microns and larger produce less oil than the smaller particles, at the same reactor conditions. This behaviour results from the finite time needed to heat the particle, which increases with

FIGURE 10 - EFFECT OF TIME ON VOLATILE YIELD



larger particle size. Correspondingly longer residence times for the larger materials should, however, produce similar yields at the identical temperatures.

5.4 Effect of Carrier Gas

Hydrogen was used as the carrier gas in Run 33. Because the test was carried out at atmospheric pressure rather than the high pressures normally associated with hydrogenation reactions, little significant change was anticipated. However, as can be seen from Figure 11, the results of this single run do show substantial improvements over tests which used nitrogen at the same conditions. Both oil and gas yields are improved by over 50%. Oil yield at 550°C is almost 40% above the Fischer Assay. Surprisingly, the unsaturated hydrocarbons have also increased. This significant improvement does need to be substantiated by additional testing, however.

The reason for the improved yield is not immediately obvious. One possibility is that hydrogen is reversing the effects of the oxidation or aging process, which is known to reduce the oil yield from shale that has been stored for lengthy periods. It should be noted that the hydrogen did not apparently react with the unsaturated gases, such as ethylene or propylene. Generally, much higher pressures are required to hydrogenate such gases.

5.5 Analysis of Oils

Only a limited analysis, Table VI, of the oils was performed. As can be seen from Figure 12, the atomic H/C ratio decreases with increasing temperature, quite the opposite effect from that observed by Watkinson. Nitrogen content increased with increasing temperature.

FIGURE 11 - EFFECT OF CARRIER GAS ON PRODUCT DISTRIBUTION

Particle Size = 600 Microns
Temperature = 550°C

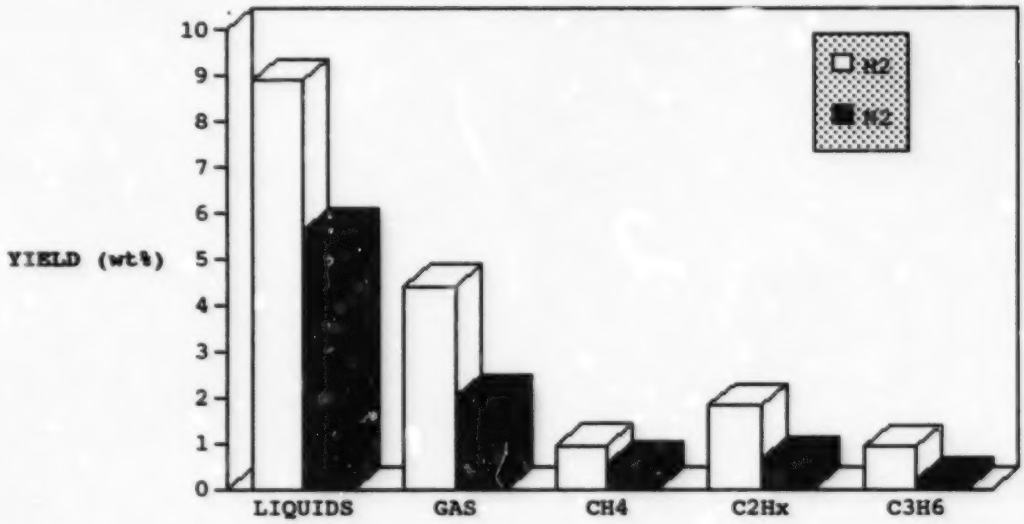
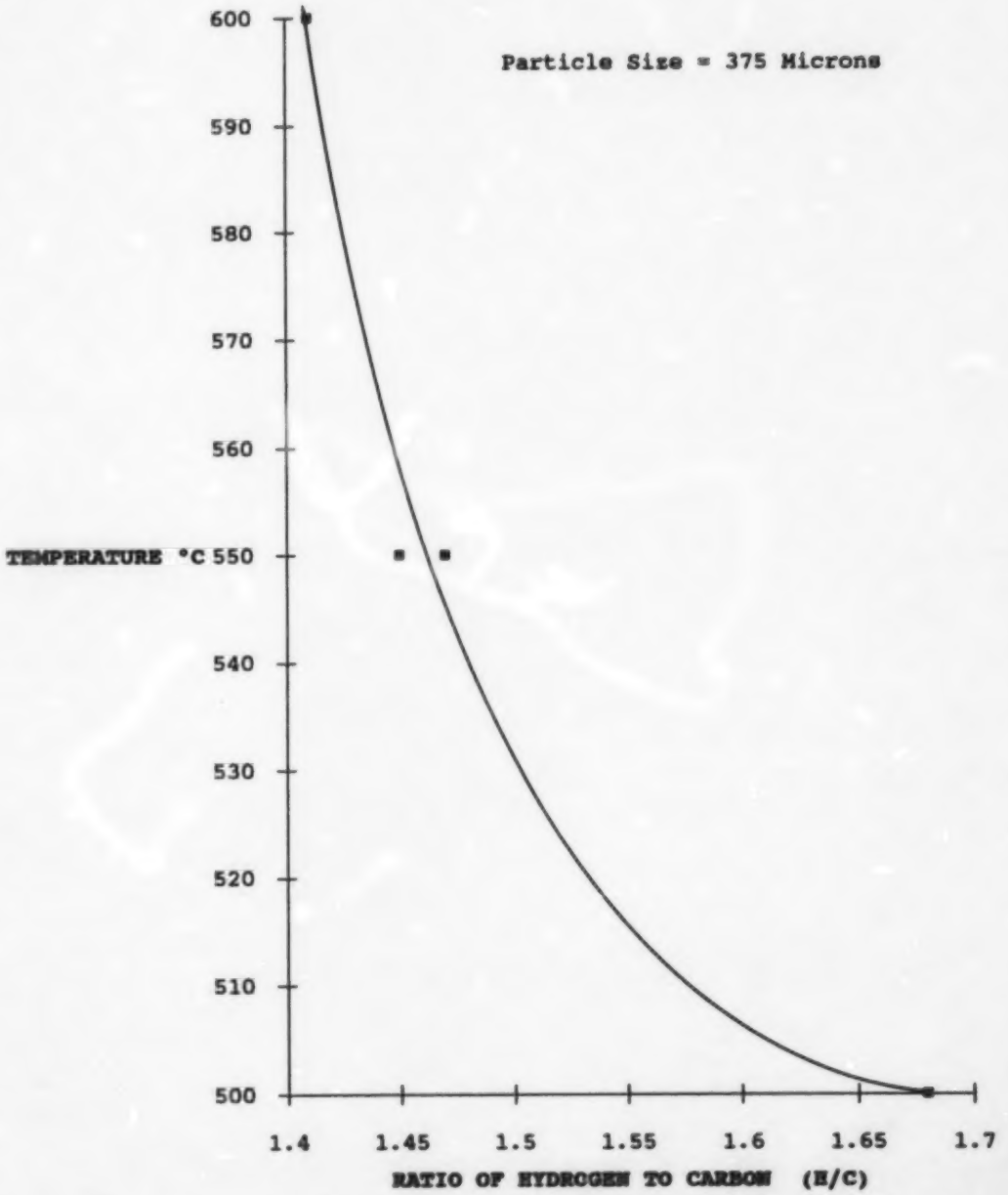


TABLE VI - EFFECT OF TEMPERATURE ON OIL ANALYSIS

Run Number	Carbon %	Hydrogen %	Nitrogen %	H/C Ratio	Temperature °C	Residue Yield
3	83.99	11.79	0.38	1.68	500	99.53
2	85.64	10.48	0.39	1.47	550	96.45
4	83.32	10.07	0.92	1.45	550	96.94
5	85.09	9.98	1.34	1.41	600	90.49

FIGURE 12 - EFFECT OF TEMPERATURE ON OIL ANALYSIS



6.0 COMMERCIAL CONSIDERATIONS

A very preliminary estimate of the economics of the process was undertaken to provide some indication of the viability of the fast pyrolysis process. The production cost of any material is primarily a function of capital cost, raw material cost, product selling price, yield, labour cost, by-product disposal cost and utilities.

To avoid the necessity of these detailed calculations, a recent report by Watts, Griffis and McOuat (1985) was used as the basis for evaluation of the process economics. All of the mining, operating and capital costs for a 20,000 T.P.D. plant were assumed to be the same as those developed in the report. No effort was made to update the costs to present day values. It was assumed that the smaller, simpler ablative reactor design would cost less than the Paraho System and that this would offset much of the annual escalation cost. In addition the mining cost would also be comparatively lower, since fines which cannot be processed in the Paraho reactor could be utilised in the ablative reactor.

The two factors which were modified, were shale yield and product selling price. These factors are shown in Table VII, together with the resultant return on investment (ROI). It can be seen from the table that at today's oil prices and utility financing, the process is marginally viable only if the yield is 40% or more above the Fischer Assay. While the results of the test program indicate that this might be possible, further process research and a more detailed cost analysis would be needed, before this could be confirmed.

TABLE VII - PRELIMINARY R.O.I. FOR A FAST PYROLYSIS OIL PRODUCTION PLANT

MATERIAL	ONT. SHALE	ONT. SHALE	N.B. SHALE	N.B. SHALE	N.B. SHALE
OIL PRICE ($US/bbl)	29	20	20	20	20
YIELD (% of F.A.)	119	119	100	120	140
OIL RECOVERY (US gal/ton	11.3	11.3	20	24	28
REVENUE ($US'000)	97839	76513	112968	129729	146489
COSTS ($US'000)	63100	63100	63100	63100	63100
DEPRECIATION ($US'000)	28050	28050	28050	28050	28050
PROFIT ($US'000)	6689	-14637	21818	38579	55339
CAPITAL ($ US'000)	561000	561000	561000	561000	561000
ROI (%)	1.2	-2.6	3.9	6.9	9.9

7.0 CONCLUSIONS

The following are the conclusions from the study on New Brunswick oil shale:

- The ablative reactor conforms to the behaviour expected of fast reactors which encompass both short vapour and solid residence times, namely, yields rapidly increase with rising temperature in the range 450 to 600°C, at residence times of less than one second. Also larger particles, in this case 850 microns, tend to produce less oil under these same operating conditions.
- At 550°C, longer residence times within the range 500 to 1200 ms improve the yield of volatiles. It is predicted from the results that a residence time of 1600 ms would be sufficient to obtain volatile yields equivalent to the Fischer Assay values, at 550°C.
- Yields of 120% and probably as high as 140% of Fischer Assay can be obtained in a fast pyrolysis reactor, such as the CAR, at operating conditions of around 650°C and residence times of 800 to 1000 ms.
- In one test, the use of hydrogen at atmospheric pressure appeared to significantly improve oil yields.
- At the highest yields obtained in the tests, 140% of Fischer Assay, the process is marginally viable at today's oil prices.

8.0 RECOMMENDATIONS

It is recommended that :

- Further testing be carried out to optimize the oil yields under conditions involving longer residence times, higher temperatures,recirculated carrier gas, hydrogen and steam.
- Develop a greater understanding of the effects of aging, oxidation and the influence of hydrogen or other reducing gases on the overall oil yield.
- A more detailed process evaluation and cost estimate be undertaken to develop a better understanding of the process viability.

9.0 LITERATURE CITED

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2. Salib P., Barua S., Abbot D., (1984) "Demonstration of an Integrated Oil Shale retort System" Canmet Report - DSS File Number 1S SQ. 23440-3-9086
3. Scott D.S., Piskorz J., (1984) "Production of Liquid from Biomass by Continuous Fast Pyrolysis" Bioenergy 84, vol 3, 15
4. Watkinson A.P., Khoe G-K., Tam T. (1985) "Rapid Pyrolysis of New Brunswick Oil Shale" Canmet Project - DSS File Number 20ST-23440-4-9000
5. Watts, Griffis and McOuat Limited (1985) "Evaluation of the Processing Options for the Oil Shales of Ontario" Canmet Project - No. 504 604 (432553)

APPENDIX A

Assessment of the Fischer Assay Analysis

Assessment of the Fischer Assay Analysis

The Fischer Assay oil yield is used as the basis of comparison for the test data from both this program and that of other workers. It is important, therefore, to determine the accuracy of this basic parameter.

From the data in Table II and Figure 2 of section 3.2, it is obvious that there is a trend towards lower oil yield as the particle size decreases. In a detailed investigation by K. E. Stanfield and I. C. Frost - U.S.B.M. Report RI 3977 (October, 1946), these investigators found no significant change in oil yield with particle size, until the particle size was less than 65 mesh. Even then, the change was less than 7%. Particle size, per se, should therefore have little effect on the oil yield. What is more likely is that the grinding, screening and storage processes have affected the oil yield of the various particle sizes.

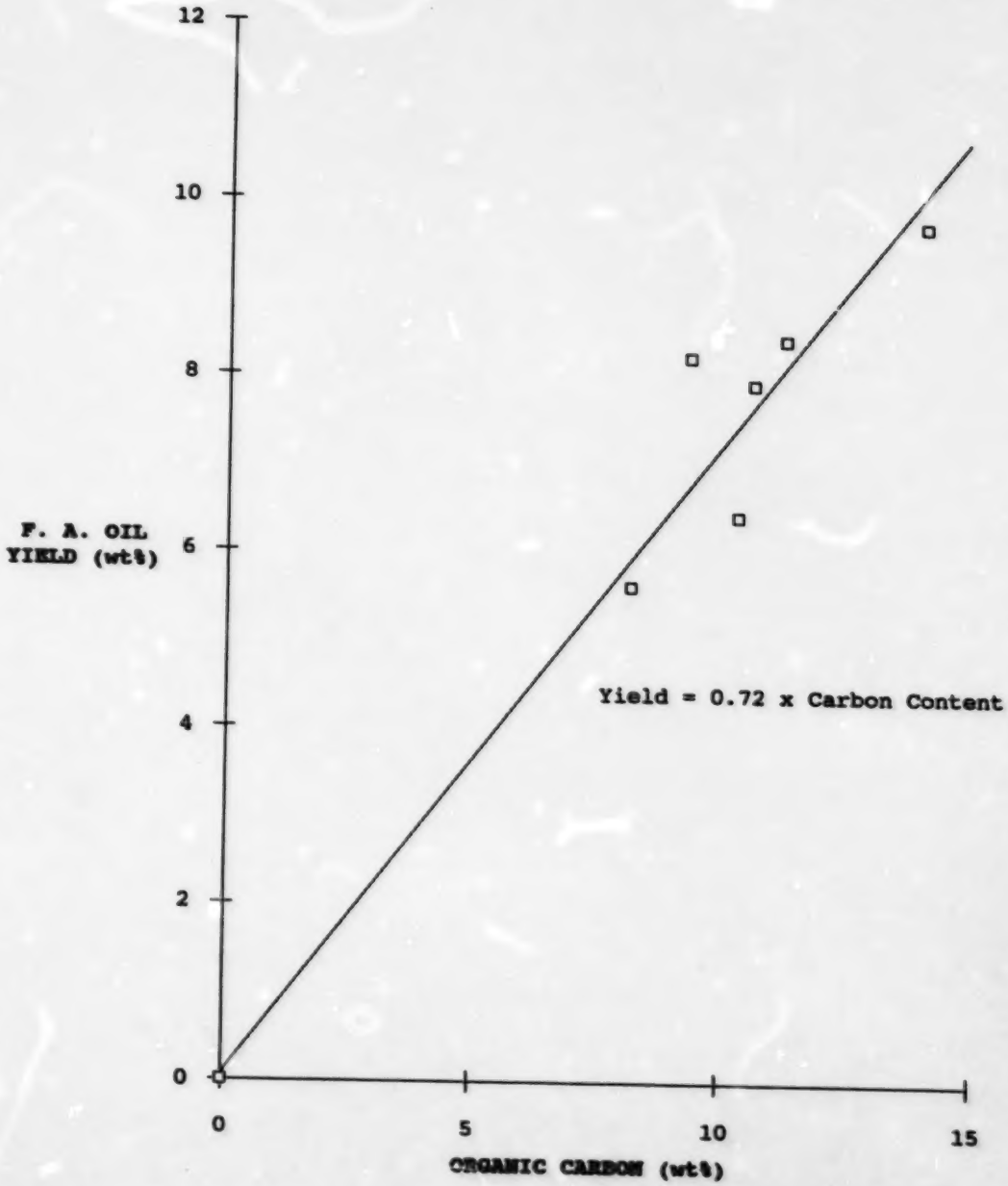
Since all of the shale has undergone the same processing and storage sequence, it would be reasonable to expect that the shale properties would vary monotonically as the particle size decreases. This is certainly true of organic carbon content and, within experimental error, of the total volatile loss from the shale, as measured by the Fischer Assay. The oil yield, however, does not obviously follow the same monotonic pattern.

As part of a test program for assessing Kentucky oil shale, Rubel et al. from the Kentucky Institute for Mining and Minerals Research analyzed a single oil shale core at approximately 3 foot intervals. The oil yield and the organic carbon content were subjected to a linear regression analysis. The data conformed well to the equation:

$$\text{Oil Yield} = 0.48 \times \text{Organic Carbon} - 0.46$$

The correlation coefficient, r^2, for the data set was 0.92. When the analysis from RPC were similarly correlated, Figure 1, the variability in oil yield was only partially explained by the change in organic carbon content, $r^2 = 0.67$.

FIGURE A1 - RELATIONSHIP BETWEEN ORGANIC CARON CONTENT AND OIL YIELD



It would appear from the above, that the measured oil yield might not be as accurate as would be implied from the Fischer Assay test procedure, i.e. true value ± 0.4. Four of the data points were within $\pm 5\%$ of the predicted value shown in Figure 1. The remaining values were -14% and +21%. If these two points are removed from the data set, the correlation remains the same, but the coefficient substantially improves, $r^2 = 0.96$.

It was not possible within the constraints of the project to carry out another set of duplicate analyses. However, the value for the +35 mesh material had already been repeated, as it was thought to be low. The initial oil yield was 2%, substantially different from the 6.4% measured in the second test. There was no obvious reason for the discrepancy between the two analyses. It would appear quite probable that there are some significant errors in the analytical oil yield for the samples used in the tests.

APPENDIX B

Detailed Table of Results

TABLE B1- RESULTS

RUN NUMBER	7	12	13	14	15	16	17
REACTOR PARAMETERS							
PARTICLE SIZE (microns)	375	375	375	375	375	375	375
TEMPERATURE oC	550	600	500	550	450	650	500
CARRIER GAS	N2	N2	N2	N2	N2	N2	N2
REACTOR PRESSURE (kPa)	114	126	139	136	143	129	134
RESIDENCE TIME (ms)	673	819	1077	893	1160	811	445
SOLIDS FLOW RATE (kg/h)	7.1	6.3	6.1	5.9	5.8	6.1	4.8
YIELD (kg)							
SHALE INPUT	9.19	4.34	6.42	8.08	6.66	6.63	2.89
RESIDUE PRODUCED	6.86	2.52	5.43	---	6.48	4.45	2.92
ASH CONTENT (wt %)	81.88	83.70	76.83	80.19	75.86	84.51	76.32
LIQUID COLLECTED	0.37	0.17	0.09	0.32	0.02	---	---
SAMPLING SYSTEM							
GAS	0.19	0.12	0.05	0.15	0.02	0.36	0.03
LIQUIDS	0.37	0.18	0.19	0.48	0.09	0.37	0.07
C6+	0.05	0.06	0.03	0.08	0.14	0.21	0.00
WATER	0.08	0.02	0.04	0.02	0.03	0.12	0.03
MASS BALANCE							
(Normalized at 100kg)							
Shale (kg)	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Shale Residue (kg)	90.47	88.50	96.42	92.38	97.65	87.65	97.06
Liquids (kg)	4.00	4.21	2.96	5.94	1.33	5.63	2.36
C6+ (kg)	0.58	1.48	0.48	0.96	2.09	3.11	0.15
Gas (kg)	2.04	2.81	0.76	1.91	0.36	5.39	0.99
Water Vapour (kg)	0.86	0.40	0.64	0.28	0.52	1.88	0.90
Unaccounted (kg)	2.05	2.61	-1.25	-1.47	-1.95	-3.65	-1.46
GAS COMPOSITION (kg)							
H2	0.019	0.004	0.001	0.009	n.d.	0.004	n.d.
CO2	0.495	0.729	0.249	0.554	0.191	0.806	0.541
CO	0.025	n.d.	n.d.	0.016	n.d.	n.d.	n.d.
CH4	0.303	0.397	0.114	0.240	0.030	0.820	0.047
C2H4+C2H6	0.562	0.851	0.151	0.473	0.066	2.043	0.078
C3H6	0.277	0.381	0.068	0.251	0.027	0.851	0.198
C3H8	0.163	0.063	0.017	0.065	0.022	0.095	0.037
C4H6	0.065	0.088	0.081	0.046	0.006	0.189	0.006
C4H8	0.090	0.176	0.049	0.138	0.011	0.345	0.054
C4H10	0.007	0.022	0.008	0.028	0.002	0.030	0.014
C5H10	0.014	0.065	0.013	0.055	0.005	0.109	0.011
C5H12	0.015	0.031	0.011	0.037	0.003	0.044	0.007
C6+	0.582	1.476	0.475	0.964	2.091	3.107	0.154
TOTAL	2.619	4.283	1.237	2.876	2.454	8.443	1.148

TABLE B1 contd. - RESULTS

RUN NUMBER	18	19	20	21	22	23	24	25
REACTOR PARAMETERS								
PARTICLE SIZE (microns)	850	850	850	850	850	850	850	850
TEMPERATURE oC	500	550	600	650	450	550	550	550
CARRIER GAS	N2	N2	N2	N2	N2	N2	N2	N2
REACTOR PRESSURE (kPa)	122	119	122	136	119	122	129	112
RESIDENCE TIME (ms)	851	748	759	802	877	536	407	1000
SOLIDS FLOW RATE (kg/h)	8.6	6.9	5.2	5.6	5.2	7.9	7.9	7.6
YIELD (kg)								
SHALE INPUT	12.59	11.55	7.51	6.37	9.08	9.25	7.99	7.98
RESIDUE PRODUCED	11.80	10.28	5.97	4.70	8.95	8.74	6.96	7.06
ASH CONTENT (wt %)	77.36	80.40	83.95	84.70	75.59	78.35	78.40	80.72
LIQUID COLLECTED	0.62	---	0.36	0.08	0.04	0.14	0.24	0.39
SAMPLING SYSTEM								
GAS	0.10	0.33	0.22	---	0.03	0.10	0.18	0.10
LIQUIDS	0.12	0.33	0.42	0.21	0.15	0.34	0.14	0.13
C6+	0.03	0.11	0.09	---	0.01	0.07	0.10	0.04
WATER	0.06	0.04	0.05	0.12	0.13	0.16	0.09	0.04
MASS BALANCE								
(Normalized at 100kg)								
Shale (kg)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Shale Residue (kg)	94.88	91.29	87.43	86.66	97.10	93.68	93.62	90.93
Liquids (kg)	0.92	2.87	5.66	3.30	1.64	3.67	1.80	1.69
C6+ (kg)	0.25	1.00	1.17	---	0.12	0.81	1.23	0.51
Gas (kg)	0.76	2.84	2.88	---	0.28	1.06	2.29	1.26
Water Vapour (kg)	0.52	0.31	0.66	1.82	1.45	1.73	1.07	0.50
Unaccounted (kg)	2.67	1.70	2.21	8.23	-0.59	-0.94	-0.01	5.11
GAS COMPOSITION (kg)								
H2	n.d.	0.012	0.013	---	n.d.	n.d.	0.003	0.005
CO2	0.099	0.688	1.030	---	0.159	0.205	0.368	---
CO	n.d.	n.d.	n.d.	---	n.d.	0.001	n.d.	n.d.
CH4	0.119	0.410	0.338	---	0.024	0.231	0.325	0.437
C2H4+C2H6	0.102	0.879	0.692	---	0.039	0.333	0.909	0.376
C3H6	0.051	0.405	0.394	---	0.020	0.118	0.318	0.164
C3H8	0.334	0.069	0.064	---	0.005	0.028	0.031	0.175
C4H6	0.011	0.117	0.111	---	0.006	0.021	0.069	0.034
C4H8	0.030	0.201	0.152	---	0.020	0.062	0.189	0.047
C4H10	0.003	0.016	0.014	---	0.000	0.011	0.010	0.004
C5H10	0.008	0.034	0.043	---	0.004	0.031	0.044	0.010
C5H12	0.004	0.014	0.025	---	0.000	0.016	0.022	0.006
C6+	0.249	0.995	1.173	---	0.123	0.807	1.234	0.510
TOTAL	1.010	3.838	4.048	---	0.400	1.865	3.523	1.767

TABLE B1 contd. - RESULTS

RUN NUMBER	26	27	28	29	30	31	32	34
REACTOR PARAMETERS								
PARTICLE SIZE (microns)	600	600	600	600	600	600	600	600
TEMPERATURE oC	550	550	500	600	650	550	550	550
CARRIER GAS	N2	N2	N2	N2	N2	N2	N2	N2
REACTOR PRESSURE (kPa)	112	112	125	125	134	127	126	125
RESIDENCE TIME (ms)	713	705	895	802	828	598	652	1094
SOLIDS FLOW RATE (kg/h)	7.3	7.3	6.5	6.5	6.5	7.1	6.3	7.2
YIELD (kg)								
SHALE INPUT	7.45	8.03	12.05	6.76	6.49	2.91	8.31	6.09
RESIDUE PRODUCED	6.52	6.96	11.34	6.00	6.03	2.43	6.58	4.98
ASH CONTENT (wt %)	80.80	80.44	76.89	83.66	84.18	81.14	79.94	83.14
LIQUID COLLECTED	0.55	0.26	0.83	0.18	0.13	0.17	0.44	0.55
SAMPLING SYSTEM								
GAS	0.16	0.16	0.10	0.26	0.48	0.17	0.99	0.28
LIQUIDS	0.36	0.29	0.30	0.41	0.37	0.24	0.92	0.27
C6+	0.11	0.10	0.06	0.14	0.20	0.21	0.12	0.10
WATER	0.03	0.04	0.11	0.03	0.10	0.04	0.12	0.04
MASS BALANCE								
(Normalized at 100kg)								
Shale (kg)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Shale Residue (kg)	92.15	92.56	96.83	89.00	88.45	91.76	93.14	89.55
Liquids (kg)	4.89	3.61	2.47	6.13	5.71	8.10	11.11	4.47
C6+ (kg)	1.53	1.20	0.47	2.04	3.08	7.19	1.43	1.66
Gas (kg)	2.10	1.99	0.80	3.85	7.41	5.99	11.90	4.58
Water Vapour (kg)	0.35	0.53	0.89	0.39	1.58	1.39	1.47	0.69
Unaccounted (kg)	-1.01	0.10	-1.45	-1.41	-6.22	-14.42	-19.05	-0.95
GAS COMPOSITION (kg)								
H2	---	---	0.000	0.003	0.029	0.054	0.026	0.079
CO2	---	---	0.227	0.723	1.394	1.088	4.934	0.522
CO	---	---	n.d.	n.d.	0.102	0.108	0.793	0.052
CH4	0.652	0.442	0.111	0.650	1.144	1.047	1.300	0.687
C2H4+C2H6	0.898	0.382	0.197	1.226	2.111	1.959	2.361	1.456
C3H6	0.163	0.231	0.120	0.507	1.246	0.896	1.308	0.928
C3H8	0.225	0.228	0.016	0.109	0.148	0.082	0.303	0.084
C4H6	0.050	0.115	0.021	0.094	0.276	0.229	0.237	0.201
C4H8	0.067	0.352	0.058	0.244	0.473	0.324	0.420	0.332
C4H10	0.008	0.124	0.007	0.044	0.279	0.035	0.052	0.030
C5H10	0.023	0.043	0.032	0.151	0.154	0.119	0.115	0.114
C5H12	0.010	0.070	0.012	0.093	0.057	0.047	0.052	0.077
C6+	1.526	1.204	0.467	2.041	3.076	7.189	1.435	1.661
TOTAL	3.623	3.192	1.270	5.885	10.489	13.177	13.336	6.223

TABLE B1 contd.- RESULTS

RUN NUMBER	1	33
REACTOR PARAMETERS		
PARTICLE SIZE (microns)	1690	600
TEMPERATURE oC	500	550
CARRIER GAS	N2	H2
REACTOR PRESSURE (kPa)	129	108
RESIDENCE TIME (ms)	435	318
SOLIDS FLOW RATE (kg/h)	15.9	6.1
YIELD (kg)		
SHALE INPUT	18.82	5.19
RESIDUE PRODUCED	---	4.61
ASH CONTENT (wt %)	76.65	82.32
LIQUID COLLECTED	0.04	0.14
SAMPLING SYSTEM		
GAS	0.14	0.23
LIQUIDS	0.04	0.38
C6+	0.05	0.08
WATER	0.05	0.16
MASS BALANCE		
(Normalized at 100kg)		
Shale (kg)	100.00	100.00
Shale Residue (kg)	98.41	90.45
Liquids (kg)	0.19	7.27
C6+ (kg)	0.28	1.62
Gas (kg)	0.76	4.41
Water Vapour (kg)	0.27	3.10
Unaccounted (kg)	0.10	-6.85
GAS COMPOSITION (kg)		
H2	0.004	n.d.
CO2	0.268	0.021
CO	n.d.	0.097
CH4	0.062	0.954
C2H4+C2H6	0.260	1.833
C3H6	0.086	0.923
C3H8	0.004	0.136
C4H6	0.025	0.128
C4H8	0.035	0.248
C4H10	0.011	0.030
C5H10	0.000	0.009
C5H12	0.002	0.032
C6+	0.276	1.622
TOTAL	1.032	6.032